Energy, Mines and Resources Canada | Énergie, Mines et Ressources Canada

CANMET

Canada Centre for Mineral and Energy Technology | Centre canadien de la technologie des minéraux et de l'énergie

ESTIMATED ENERGY REQUIREMENTS FOR THE RECOVERY OF HYDROCHLORIC ACID AND THE REJECTION OF IRON FROM AQUEOUS SOLUTIONS CONTAINING HYDROCHLORIC ACID AND FERROUS CHLORIDE

F.J. Kelly
Office of Technology Transfer

March 1984

MINERALS RESEARCH PROGRAM
MRP/OTT 84-01

36 pp

TABLE OF CONTENTS

Page

1. Introduction.....
2. General.....
3. Lurgi Process.....
4. PORI Process.....
5. Discussion.....
6. Conclusion.....
7. References.....

Table 1 Coal Analysis.....
Table 2 Heavy Fuel Oil.....
Table 3 GCM Copper Process Hydrochloric Acid Recovery Estimated Energy Requirement.....

Figure 1 Lurgi HCl Recovery System, Coal, Material and Energy Balance.....

Figure 2 Lurgi HCL Recovery System, Natural Gas, Material and Energy Balance.....

Figure 3 PORI HCl Recovery System, Heavy Fuel Oil, Material and Energy Balance.....

Figure 4 PORI HCl Recovery System, Natural Gas, Material and Energy Balance.....

8. Appendix .....

1. INTRODUCTION

A problem area in the Great Central Mines copper process [1] is the prevention of the buildup of iron and water within the circuit. In the present flowsheet this function is accomplished by crystallizing hydrated ferrous chloride from a copper free catholyte bleed stream. Water is removed by dehydrating the crystals in a dryer. The dehydrated crystals are then decomposed in the presence of hydrogen. The off gases and the crystallizer vapour are passed through an absorber and the recovered hydrochloric acid is recycled to the process. The sponge iron by-product is recovered for market.

An alternative to the above method for iron removal and Hcl recycle is to oxidize the ferrous chloride bleed stream with air to produce a discard iron oxide product.

Two processes used by the steel industry to recover hydrochloric acid from waste pickle liquor streams could be used to accomplish this. One process using fluidized-bed combustion, evaporation and acid absorption was developed by Lurgi [2,3]. The other process using evaporation, oxidation, hydrolysis and acid absorption was developed by PORI [4,5]. This report compares the estimated energy requirements to utilize these processes in the Great Central Mines process flowsheet, and indicates which process is likely to have the lowest capital and operating costs.

2. GENERAL

The composition, component weight and material balances for streams flowing to or from process equipment units in the following paragraphs are given in the appendix. The basis for all material flows is related to Great Central Mines' incremental process feed unit of 90.72 tonne (100 short tons) of dry concentrate per day. The input stream for the hydrochloric acid recovery processes (termed Weak $FeCl_2$ Liquor in the appendix) is the crystallizer feed stream (1).

The base temperature for all heat or energy calculations is 25°C. Heats are calculated from theoretical heats of formation, i.e., reaction heat = heat of formation of products - heat of formation of reactants. The sensible heat content of all streams entering or leaving the system is calculated from heat capacity data with reference to 25°C and the temperature specified for each stream. The net heat needed or rejected by the system is then determined, i.e., net heat = reaction heat + sensible heat out - sensible heat in. A negative net heat quantity indicates that operation of this system will produce an output of energy. A positive net heat quantity indicates that operation of this system will require an input of energy.

All chemical reactions are assumed to go to 100% completion in the calculation of reaction and product weights.

For both the Lurgi and PORI processes it is assumed that the composition of the gas stream entering the falling film condenser absorber unit used in the PORI process [5] contains the specified composition, combustion gases excluded, 30 wt % HCl and 70 wt % H_2O so that the condensed stream contains 30-31 wt % HCl. This allows both systems to be compared using identical starting and endpoint conditions. For simplicity it is also assumed that all HCl is vaporized in the combustor of the Lurgi process and the hydrolyzer vessel of the PORI process.

3. LURGI PROCESS

The Lurgi hydrochloric acid recovery process [2,3] consists of the following major equipment items;

- o venturi-evaporator,
- o fluidized-bed combustor-reactor,
- o cyclone,
- o acid absorber, and
- o scrubber.

When operating; weak $FECl_2$ liquor flows to the venturi-evaporator where it is contacted with the hot exhaust gases from the reactor to evaporate a portion of the contained water. Concentrated $FECl_2$ liquor from the evaporator is then pumped to the fluidized-bed reactor where it decomposes at 800°C into water vapour, hydrogen chloride gas and solid Fe_2O_3. The exhaust gases pass through a cyclone to separate solid particles carried over from the reactor before going to the evaporator. Solids collected in the cyclone are recycled to the reactor. The cooled gases leaving the evaporator are passed through a hydrochloric acid absorption tower, then a tail gas scrubber and exhausted to the atmosphere. Solid Fe_2O_3 overflowing the reactor is either sold or sent to a disposal area and recovered acid is recycled to the process.

An industrial installation using the Lurgi system is operated by Dominion Foundries and Steel Ltd. [3] (Dofasco), Hamilton, Ontario. The plant consists of four identical units, outlined above, that treat a waste pickle liquor stream containing up to 19 to 20 kg/min iron and 10

to 11 kg/min hydrochloric acid. The fluidized-bed combustors burn natural gas. A similar experimental system being tested in Australia [6] burns coal in the fluidized-bed combustor.

Before making an energy balance around the fluidized-bed combustor an estimate of the weight of the fluidizing air will be required. Also needed will be an estimate of the weight of combustion products produced by the burning of the fuels, coal or natural gas.

Assumptions made in estimating the size of the fluidized-bed area are as following:

a) average of the iron feed rate used at the Dofasco plant 1.7 $kg/m^2/min$;

b) average fluidizing air flowrate used in the Lurgi and Australian combustors of 17.07 $m^3/m^2/min$ at 0°C and 1 atm.

For a plant treating 90.72 tonnes of dry concentrate per day the iron flowrate in the HCl recovery stream is 18.95 kg/min.

$$\text{Combustor area} = \frac{\text{iron flow}}{\text{iron flow per unit of area}}$$

$$= \frac{18.95}{1.7}$$

$$= 11.15\ m^2$$

$$\text{Combustor diameter} = \sqrt{\frac{4 \times \text{area}}{\pi}}$$

$$= \sqrt{\frac{4 \times 11.15}{3.1416}}$$

$$= 3.77 \text{ m}$$

Fluidizing air = bed area x air flowrate per unit area

$$= 11.15 \times 17.07$$

$$= 190.33 \text{ m}^3/\text{min}$$

$$\text{Weight of air} = \frac{\text{volume x mole weight air}}{\text{volume/kg mole}}$$

$$= \frac{190.33 \times 28.97}{22.41}$$

$$= 245.7 \text{ kg/min} = 353.8 \text{ tonne/day}$$

An analysis of a typical bituminous coal is given in Table 1 and that for heavy fuel oil in Table 2. Natural gas is assumed to be 100% methane with a higher heating value of 37.26 MJ per m^3 at 0°C and 1 atmosphere. In estimating the higher heating value of coal it was assumed that;

a) 6% of the carbon content is not burned [6];
b) contained oxygen is tied up as water;
c) heat taken up by evaporation of contained and reaction product water, heating up of reaction product gases and radiation heat loss amounts to 30% of the available heat from coal, natural gas and heavy fuel oil.

Main chemical reactions occuring within the Lurgi system are as follows:

$$4FeCl_2 + O_2 \rightarrow 4H_2O + 2Fe_2O_3 + 8HCl \qquad \text{Eq 1}$$

$$H_2O \text{ (liquid)} \rightarrow H_2O \text{ (vapour)} \qquad \text{Eq 2}$$

$$C + O_2 \rightarrow CO_2 \qquad \text{Eq 3}$$

$$2H_2 + O_2 \rightarrow 2H_2O \qquad \text{Eq 4}$$

$$S + O_2 \rightarrow SO_2 \qquad \text{Eq 5}$$

The higher heating value (HHV) of the dry coal was estimated as follows:

a) **Carbon available for combustion**

$C = 78.88 - (0.06 \times 78.88) = 74.15\%$

b) **Hydrogen available for combustion**

$H_2 = 4.42 - (6.55/8) = 3.6\%$

c) **Locked in moisture**

$H_2O = 6.55 + (6.55/8) = 7.37\%$

d) **Higher heating value**

$C = 0.7415 \times 34.64 = 25.68$ MJ/kg

$H = 0.036 \times 142.02 = 5.11$ MJ/kg

$S = 0.004 \times 9.26 = 0.04$ MJ/kg

Coal HHV = 30.83 MJ/kg (13265 Btu/lb)

e) **Lower heating value (LHV)**

LHV = HHV - 30% HHV

LHV = 30.83 - (0.3 x 30.83) = 21.58 MJ/kg (9285 Btu/lb)

f) **Heat duty is estimated to be 585 GJ/d**

$$\text{Estimated coal} = \frac{585. \times 10^{11}}{21.85 \times 10^{6}} = 26774 \text{ kg/d}$$

approx. 26.8 dry t/d = 29.2 moist t/d

Using the above technique the higher heating value of the heavy fuel oil was estimated to be 40.97 MJ/kg (17627 Btu/lb) with 7% of the carbon not burned.

Using the molar ratios shown in equations 3, 4 and 5, the theoretical combustion air and gaseous combustion products were estimated as follows:

a) **Air**

C = 0.7415 x 2.67 = 1.98

H_2 = 0.036 x 8 = 0.29

S = 0.004 x 1 = 0.004

O_2 Total = 2.27 kg/kg of coal

$$\text{Stiochmetric air} = \frac{2.274}{0.232} = 9.8 \text{ kg/kg dry coal}$$

b) **Gaseous Products**

CO_2 = 0.7415 x 3.67 = 2.72 kg/kg dry coal

SO_2 = 0.004 x 2 = 0.008 kg/kg dry coal

H_2O = 0.036 x 9 = 0.324 kg/kg dry coal

N_2 = 0.768 x 9.8 = 7.53 kg/kg dry coal

Process stream components shown in the appendix related to the combustion process were calculated with the above factors. The operating conditions, combined energy and overall material balances for the adapted Lurgi process using coal is shown in Table 3 and Figure 1 and Table 3 and Figure 2 using natural gas.

4. PORI PROCESS

The PORI hydrochloric acid recovery process [4,5] consists of the following major equipment items; venturi-evaporator, oxidizer, venturi-hydrolyzer, acid absorber and scrubber. When operating; weak $FeCL_2$ liquor flows to the venturi-evaporator where it is contacted with hot burner exhaust gases to evaporate a portion of the contained water. Concentrated $FeCl_2$ liquor from the evaporator is then pumped to the oxidation pressure vessel, which operates at 690 kPa and 149°C, where $FeCl_2$ is converted to $FeCl_3$ and Fe_2O_3 (Reaction 6). The slurry from the oxidizer is pumped to the venturi-hydrolyzer operating at 316°C. In the hydrolyzer aqueous $FeCl_3$ solution is contacted with hot burner exhaust gases wnich evaporate the water and decompose $FeCl_3$ into HCl gas and solid Fe_2O_3 (Reaction 7). The exhaust gas from the hydrolyzer is combined with the oxidizer exhaust gas and passed through a hydrochloric acid absorption tower, then a tailgas scrubber and exhausted to the atmosphere. Solid Fe_2O_3 is either sold or sent to a disposal area and recovered hydrochloric acid is recycled to the process.

An industrial installation using the PORI system is operated by J&L Steel Corporation [5], Cleveland, Ohio. The plant treats a waste pickle liquor stream flowing at 225 liters per minute and containing 27.21 kg $FeCl_2$ (12 kg Fe^{++}). Heat supplied to the venturi-evaporator is supplied by auxiliary burners fired with natural gas or heavy fuel oil which are the fuels used in this application.

The main chemical reactions occuring in the PORI process are as follows:

Oxidizer; $12FeCl_2 + 3O_2 \rightarrow 8FeCl_3 + 2Fe_2O_3$ Eq 6

Hydrolysis; $2FeCl_3 + 3H_2O \rightarrow 6HCl + 2Fe_2O_3$ Eq 7

Estimating the heating value of heavy fuel oil and natural gas and the weights of combustion air, plus excess over stoichiometric, and combustion gas products was done as outlined in the Lurgi process. An estimate of air flow weight, 50% in excess of stoichiometric, and the amount of vapour exhausting with the oxidizer gas was made as follows:

a) Vapour pressure H_2O at 149°C is 461.8 kPa

b) From reaction 6 the stoichiometric O_2 is 3.91 t/d

$$\text{Air} = 1.5 \times \frac{3.91}{0.232} = 25.28 \text{ t/d} = 13.58 \text{ m}^3/\text{min}$$

c) Air left after reaction assuming remaining gas is air

Air = 11.7 m^3/min

$$\text{Vol (689.3 kPa, 149°C)} = \frac{422\ (101.3)(11.7)}{273\ (790.6)} = 2.32 \text{ m}^3/\text{min}$$

d) Water evaporated at 0°C and 101.3 kPa and H_2O vol = Air vol

$$H_2O = \text{Vol} \times \frac{273}{422} \times \frac{VPH_2O}{AtmP} \times \frac{MWH_2O}{\text{kg-mole}} \times \frac{\text{kg-mole}}{m^3}$$

$$H_2O = 2.32 \times \frac{273}{422} \times \frac{461.8}{101.3} \times \frac{18.015}{22.41} = 5.5 \text{ kg/m} = 7.92 \text{ t/d}$$

Process stream components shown in the appendix related to the combustion process and other physical relationships were calculated with the above factors. The operating conditions, combined energy and material balances for the adapted PORI process using heavy fuel oil is shown in Table 3 and Figure 3 and Table 3 and Figure 4 using natural gas.

5. DISCUSSION

The PORI process requires less energy input to recover hydrochloric acid from solution than does the Lurgi process (Table 3). Without a heat recovery step in the Lurgi process, the energy difference is 220.7 GJ per feed unit (90.72 tonnes dry concentrate) when the PORI process is gas fired. This quantity of energy is equivalent to 7.9 tonnes of dry coal. If heat recovery is added to the Lurgi system, the energy difference between it and the PORI system is reduced to 42.5 GJ or 1.4 tonnes of dry coal. However the heat recovery step will require the purchase and insertion of a heat exchanger within the process. A use for the steam product will also be required.

The Lurgi process also requires more air, a difference of 121 tonnes per feed unit (Figures 1 and 4), than does the PORI process. However, 25.3 tonnes of air per feed unit must be supplied to the oxidation vessel at a pressure of 689.3 kPa. This also means that high pressure pumps will be required to feed solution into the oxidation vessel. Prior to entering the oxidation vessel, the temperature of the feed solution is raised 45°C. This will mean the purchase of a heat exchanger, mentioned in the process description (4), but not shown in the flow diagram (4,5). Another requirement of the oxidation step is that completion of the oxidation reaction (Eq 6) requires a solution retention period in the oxidation vessel of two hours (5). This factor will have an effect on determining the size of the agitated oxidation pressure vessel.

According to the Australian paper [6] a disadvantage of the fluidized-bed process is that it is limited to the use of light fuel oils and gaseous fuels. Heavy fuel oil causes combustion problems that lead to defluidization of the bed. Even when burning light fuel oils or gaseous fuels, the outside diameter of the combustor is limited to 3 m[6]. At the Dofasco plant, the Lurgi fluidized-bed combustors are 3 m outside diameter. On the other hand, there should be no restriction on the diameter of coal burning fluidized-bed combustors, since 7 to 9 m diameter combustors are presently used in electric power generation plants. Using the average Dofasco combustor feed rate of 1.7 kg $Fe^{++}/m^2/min$ of bed area and feed units, a total plant feed rate of 635 tonnes dry concentrate per day, the total combustor area required is 78 m^2. This is equivalent to twenty parallel process streams containing 20 combustors, 3 m diameter, burning gaseous fuel or light fuel oil or two parallel process streams containing 2 combustors, 8 m diameter, burning coal. The twenty parallel systems will probably require a floor area of 2000 m^2 and the two coal burning systems 800 m^2. To this area, more would have to be added to accomodate utilities, motor control centres, control room, laboratory, workmen's facilities, etc.

The estimated PORI process design flow rate for the plant at J&L Steel Corp. [5] is 27.22 kg Fe^{++}/min. The total floor area of the plant is approximately 560 m^2. Half of this space is taken up by the process equipment which has 100% installed spare capacity. Therefore it is assumed that the required floor area would be approximately 140 m^2 to accomodate the equipment for the above flow rate. One dry concentrate unit contains 1.58 times the above Fe^{++} flow rate. The Fe^{++} flow rate produced by the plant at full production will be eleven times greater than the PORI design Fe^{++} flow rate. Assuming that the PORI process can

accommodate a ten fold increase in flow and using a scale-up exponent of 0.63, the estimated floor area for the process is:

Area = $140(11)^{0.63}$ = 635 m^2 say 700 m^2. Or two parallel units half this size in 850 m^2.

6. CONCLUSIONS

An estimate of the energy required to remove iron and water and recover hydrochloric acid from the Great Central Mines hydrometallurgical process has been determined. The processes investigated to perform this task were the Lurgi fluidized-bed combustion system and the PORI oxidation-hydrolization system. The following is a comparison of the relative requirements of both systems:

1. Firing of the Lurgi fluidized-bed combustors with oil or gas requires a large number of small parallel units due to a limitation on the combustor diameter. The plant size, and capital and operating costs are therefore likely to be up to a factor of two higher than for the PORI process.

2. Capital costs of a coal fired Lurgi system could be competitive with the PORI process.

3. The coal fired Lurgi system without heat recovery requires 727 GJ (690 MBTU) per feed unit (90.72 metric tonnes or 100 short tons dry concentrate), or 28 MJ/kgCu (12100 BTU/lb Cu).

4. With heat recovery the coal fired Lurgi system requires 549 GJ (521 MBTU) per feed unit, or 21 MJ/kg Cu (9050 BTU/lb Cu).

5. The natural gas fired PORI system requires 506 GJ (481 MBTU) per feed unit, or 20 MJ/kg Cu (8450 BUT/lb Cu).

6. The PORI system has lower operating temperatures 149-177°C when compared to the 850°C in the Lurgi combustor.

6. The operating pressure in the Lurgi system is 103 kPa while that in the PORI system ranges from 103 to 690 kPa.

7. Air requirements in the Lurgi system are 354 tonnes per feed unit and that of the PORI system 233 tonnes per feed unit. Eleven percent of the PORI air requires compression to 690 kPa.

7. REFERENCES

1. Craigen, W.J.S. and Kelly, F.J., 'Great Central Mines Hydrometallurgical Process - Evaluation of Modified Process and Recommendations for Commercial Application', MRP/OTT 82-01 (CF), CANMET, Energy, Mines and Resources, Canada, November, 1982.

2. 'Continuous Regeneration of Hydrochloric Acid Pickle Liquors', Lurgi Express Information Bulletin T1032/4.72, Dampftechnik Division, Lurgi Apparate - Technik GmbH, Frankfurt (Main), POB 119181.

3. Rupay, G.H. and Jewell, C.J., 'The Regeneration of Hydrychloric Acid from Waste Pickle Liquor Using the Keramchemie/Lurgi Fluidized-Bed Reactor System', CIM Bulletin, February, 1975.

4. Burtch, J.W., 'PORI Hydrochloric Acid Regeneration Process', Iron and Steel Engineer, April, 1973.

5. Burtch, J.W., 'Hydrochloric Acid from Industrial Waste Streams - The PORI Process', CIM Bulletin, January, 1975.

6. Sinha, H.N. and Tuftley, J.R., 'Recovery of Hydrochloric Acid from Ferrous Chloride Liquor in a Coal-Fired Fluidized-Bed', CSIRO Division of Mineral Chemistry, Port Melbourne and Australian Mineral Development Laboratories, Frewville, Australia.

TABLE 1 COAL ANALYSIS

	Dry Basis Weight (%)
Carbon	78.88
Hydrogen	4.42
Nitrogen	1.05
Sulphur	0.40
Oxygen	6.55
Ash	8.70
Total	**100.00**
Moisture (wet basis)	8.00

TABLE 2 HEAVY FUEL OIL

	Weight %
Carbon	88.6
Hydrogen	10.0
Sulfur	1.0
Oxygen	0.4
Total	**100.0**

API gravity 12.; Specific gravity 0.9861

TABLE 3 GCM COPPER PROCESS HYDROCHLORIC ACID RECOVERY
Estimated Energy Requirements

PROCESS AREA	GJ/90.72 t Dry Concentrate	MJ/kg Copper
Lurgi Process Coal	727.33	28.11
Lurgi Process Coal with Heat Recovery	548.80	21.21
Lurgi Process Natural Gas	688.16	26.60
Lurgi Process Natural Gas with Heat Recovery	506.30	19.57
PORI Process no. 6 Fuel Oil	505.11	19.52
PORI Process Natural Gas	506.63	19.58

M: mega G: giga J: joule t: tonne

lb = 0.4536 kg
MBTU = 1.054 GJ
short ton = 0.9072 t

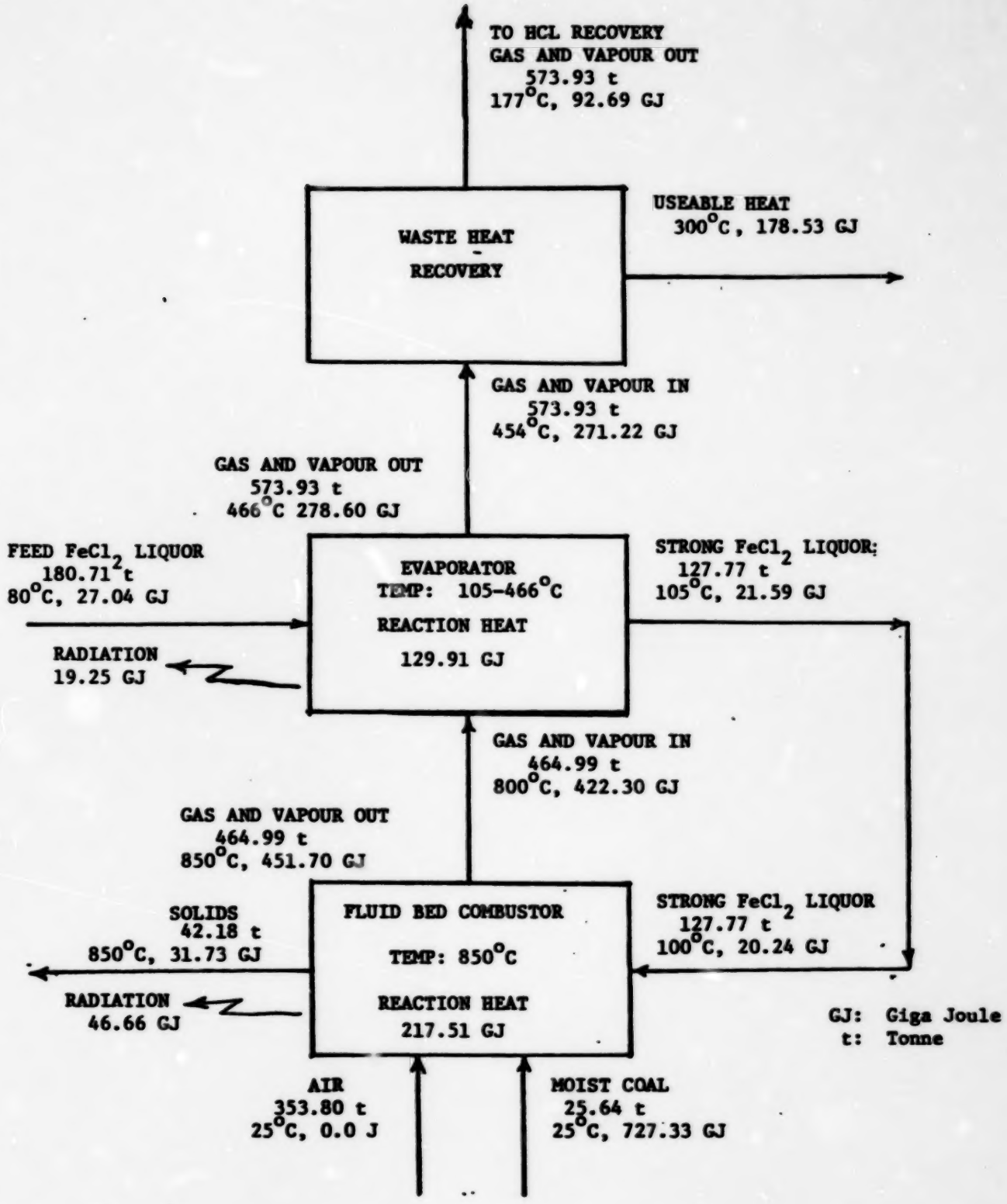

FIGURE 1: LURGI HCL RECOVERY SYSTEM, COAL, MATERIAL AND ENERGY BALANCE

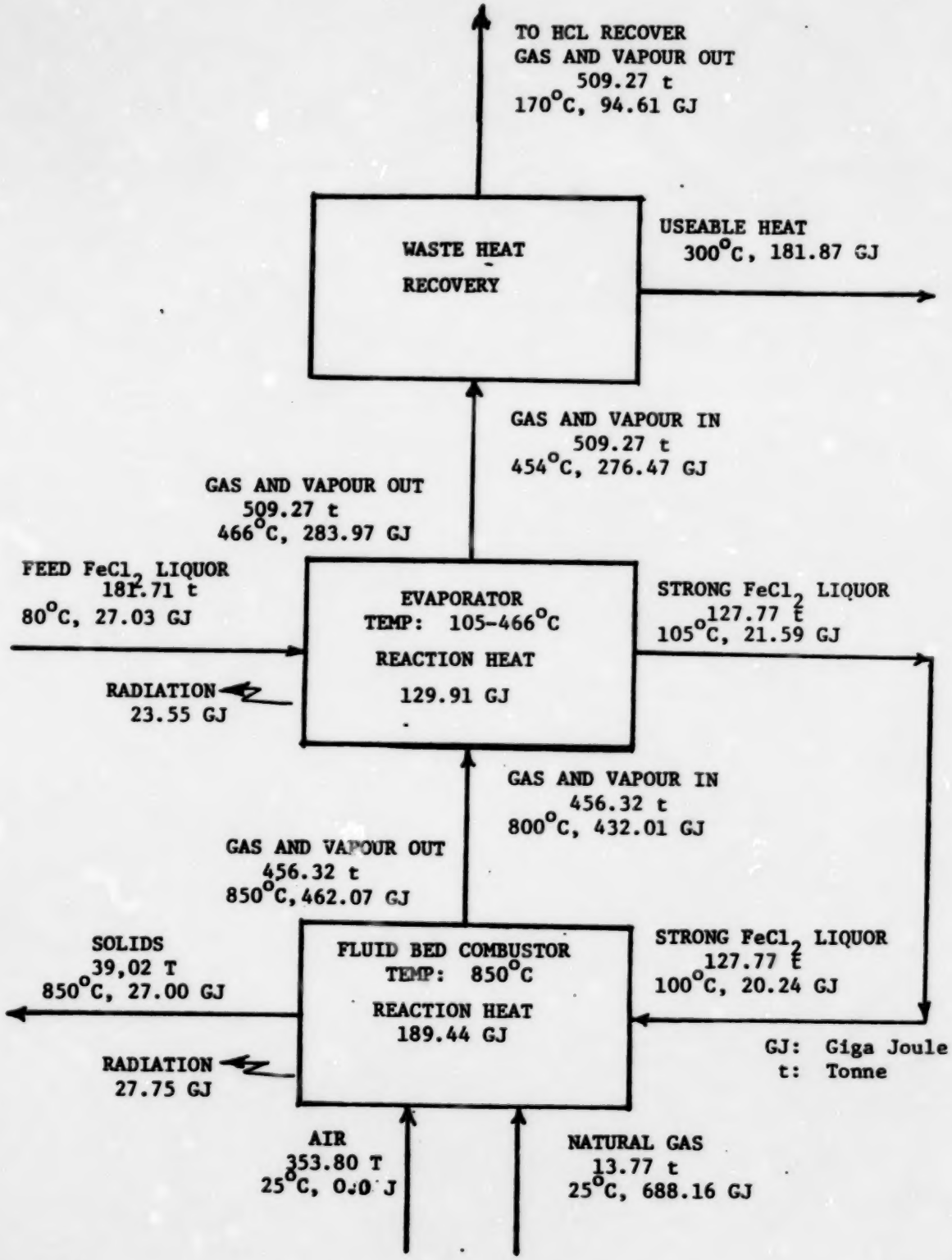

FIGURE 2: LURGI HCL RECOVERY SYSTEM, NATURAL GAS, MATERIAL AND ENERGY BALANCE

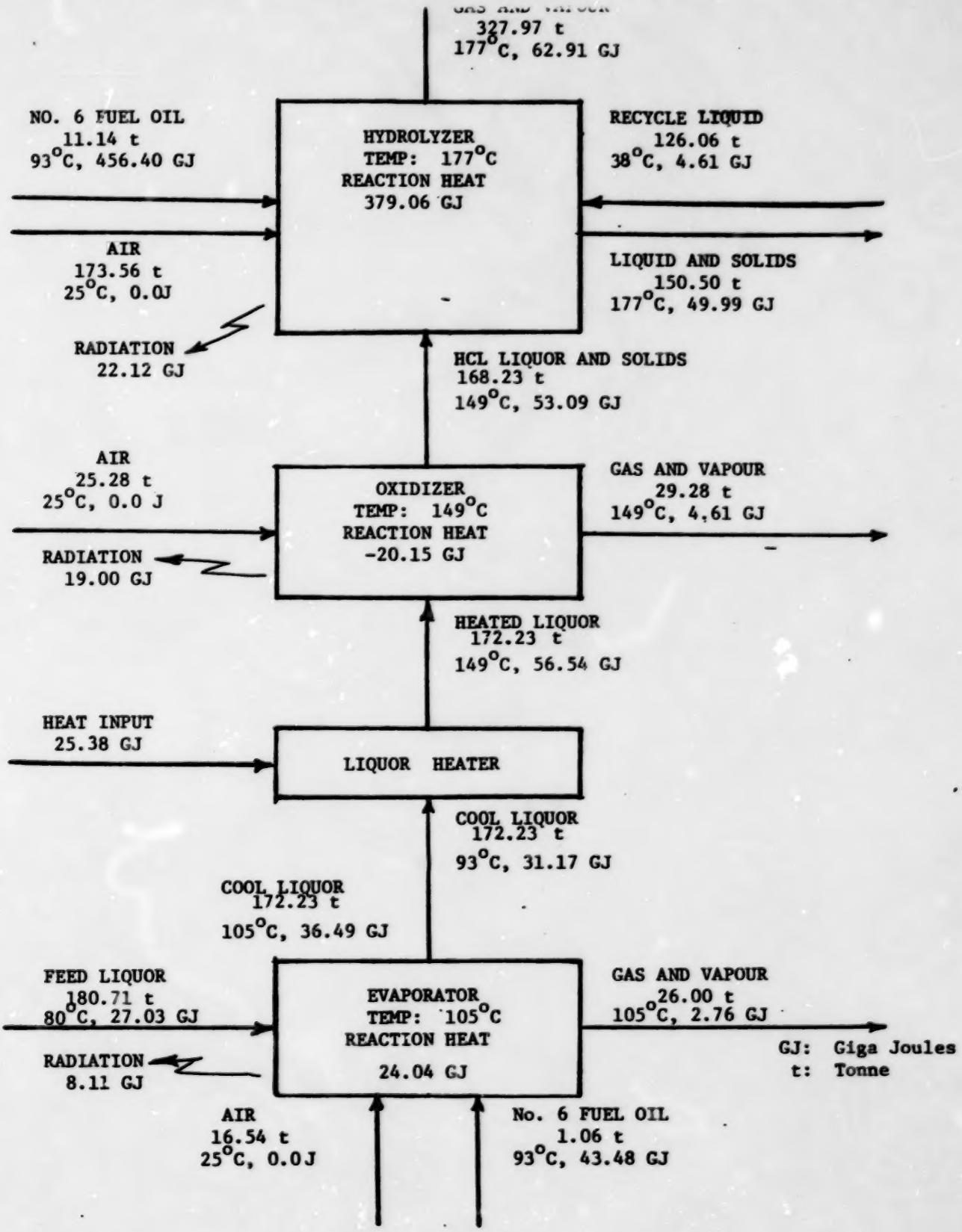

FIGURE 3: PORI HCL RECOVERY SYSTEM, HEAVY FUEL OIL, MATERIAL AND ENERGY BALANCE

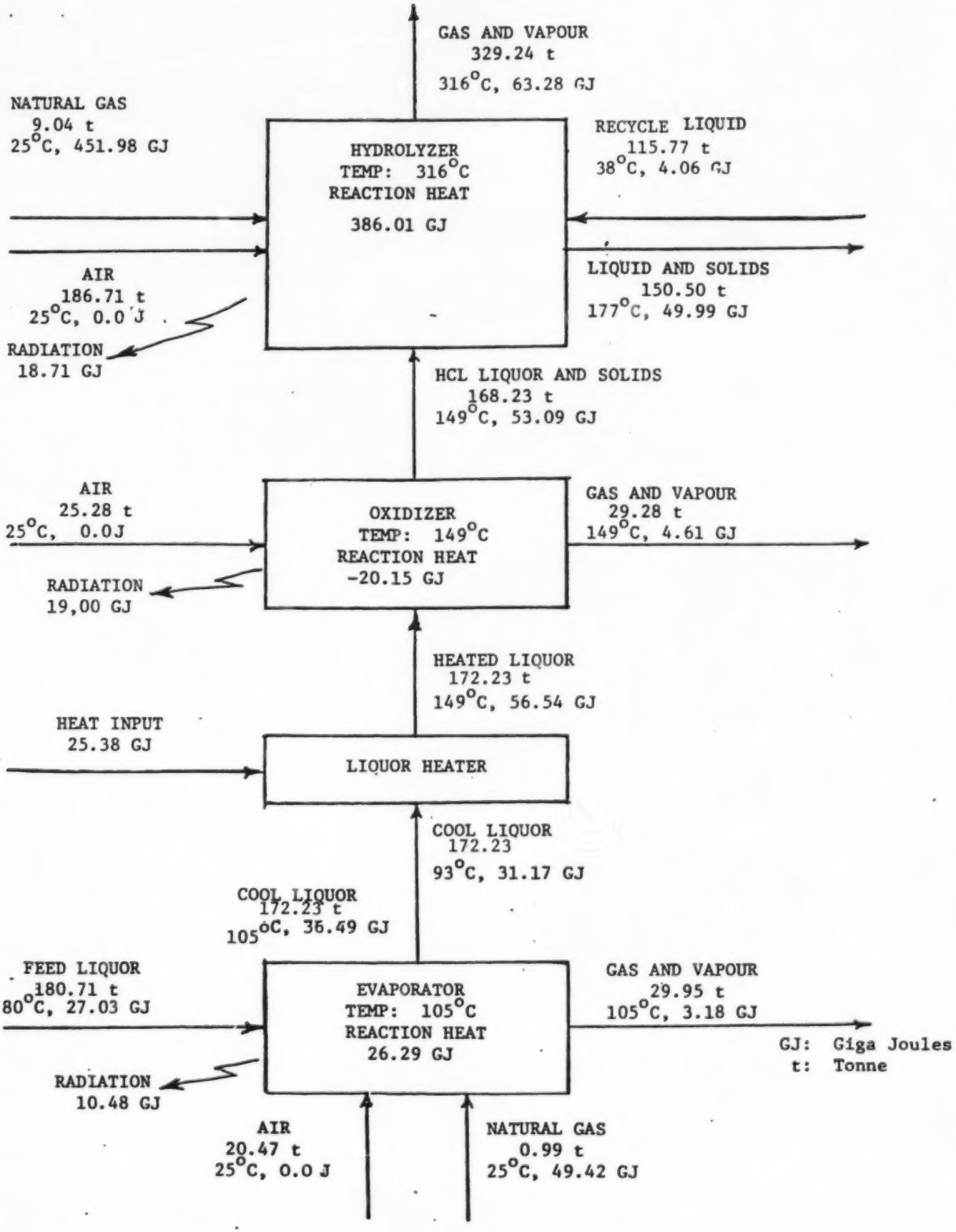

FIGURE 4: PORI HCL RECOVERY SYSTEM NATURAL GAS MATERIAL AND ENERGY BALANCE

8. APPENDIX

Lurgi and PORI Process Stream Contents in tonnes per 90.72 tonnes of dry concentrate.

Symbols

AQ : aqueous solution
G : gas
L : liquid
S : solid
m : Meter
kg : kilogramme
M : mega 10^6
G : giga 10^9
J : joule
BTU: British thermal unit
kPa: kilo Pascals
min: minute
d : day
t : tonne

Conversion Factors

Unit	Times	To Unit
kilogramme	2.2046	pounds
tonne	0.9072	short ton
joule	0.000948	British thermal unit
meter	3.28084	feet
kilo Pascal	0.009872	atmosphere

Lurgi $FeCl_2$ fluidized-bed combustor using coal

		Tonnes
Coal in		
Coal		23.59
H_2O-L		2.05
	Total	**25.64**
Air-G in		353.80
Strong $FeCl_2$ liquor in		
Hcl-AQ		12.89
H_2O-L		52.94
$FeCl_2$-AQ		61.93
	Total	**127.76**
	Total in	**507.20**
Gas and vapour out		
Air-G		103.59
N_2-G		192.91
HCl-G		48.53
CO_2-G		64.19
SO_2-G		0.19
H_2O-G		55.57
	Total	**464.98**
Solids out		
Fe_2O_3-S		39.02
C-S		1.12
Ash-S		2.05
	Total	**42.19**
	Total out	**507.17**

Lurgi $FeCl_2$ evaporator using coal

		Tonnes
Weak $FeCl_2$ liquor in		
HCl-AQ		12.89
H_2O-L		105.89
$FeCl_2$-AQ		61.93
	Total	**180.71**
Gas and vapour in		
Air-G		103.59
N_2-G		192.91
HCl-G		48.53
CO_2-G		64.19
SO_2-G		0.19
H_2O-G		55.57
	Total	**464.98**
	Total in	**645.69**
Strong $FeCl_2$ liquor out		
HCl-AQ		12.89
H_2O-L		52.94
$FeCl_2$-AQ		61.93
	Total	**127.76**
Gas and vapour out		
Air-G		103.59
N_2-G		192.91
HCl-G		48.53
CO_2-G		64.19
SO_2-G		0.19
H_2O-G		108.51
	Total	**517.92**
	Total out	**645.68**

Lurgi $FeCl_2$ fluidized-bed combustor using natural gas

		Tonnes
Gas in		
Air-G		353.80
Natural gas		13.77
	Total	**367.57**
Strong $FeCl_2$ liquor in		
HCl-AQ		12.89
H_2O-L		52.94
$FeCl_2$-AQ		61.93
	Total	**127.76**
	Total in	**495.33**
Gas and vapour out		
Air-G		43.05
N_2-G		251.92
HCl-G		48.53
H_2O-G		75.06
CO_2-G		37.78
	Total	**456.34**
Solids out		
Fe_2O_3-S		39.02
	Total out	**495.36**

Lurgi $FeCl_2$ evaporator using natural gas

		Tonnes
Weak $FeCl_2$ liquor in		
HCl-AQ		12.89
H_2O-L		105.89
$FeCl_2$-AQ		61.93
	Total	**180.71**
Gas and vapour in		
Air-G		43.05
N_2-G		251.92
HCl-G		48.53
H_2O-G		75.06
CO_2-G		37.78
	Total	**456.34**
	Total in	**637.05**
Strong $FeCl_2$ liquor out		
HCl-AQ		12.89
H_2O-L		52.94
$FeCl_2$-AQ		61.93
	Total	**127.76**
Gas and vapour out		
Air-G		43.05
N_2-G		251.92
HCl-G		48.53
H_2O-G		128.00
CO_2-G		37.78
	Total	**509.28**
	Total out	**637.04**

PORI $FeCl_2$ evaporator using no. 6 fuel oil

		Tonnes
Energy in		
Air-G		16.54
No. 6 fuel oil		1.06
	Total	**17.60**
Weak $FeCl_2$ liquor in		
HCl-AQ		12.89
H_2O-L		105.89
$FeCl_2$-AQ		61.93
	Total	**180.71**
	Total in	**198.31**
Strong $FeCl_2$ liquor out		
HCl-AQ		12.89
H_2O-L		97.41
$FeCl_2$-AQ		61.93
	Total	**172.23**
Gas and vapour out		
CO_2-G		3.22
O_2-G		0.64
N_2-G		12.70
H_2O-G		9.43
	Total	**25.99**
	Total out	**198.22**

PORI $FeCl_2$ evaporator using natural gas

		Tonnes
Energy in		
Air-G		20.47
Natural gas		0.99
	Total	**21.46**
Weak $FeCl_2$ liquor in		
HCl-AQ		12.89
H_2O-L		105.89
$FeCl_2$-AQ		61.93
	Total	**180.71**
	Total in	**202.17**
Strong $FeCl_2$ liquor out		
HCl-AQ		12.89
H_2O-L		97.41
$FeCl_2$-AQ		61.93
	Total	**172.23**
Gas and vapour out		
CO_2-G		2.72
O_2-G		0.79
N_2-G		15.72
H_2O-G		10.71
	Total	**29.94**
	Total out	**202.17**

PORI oxidation reactor

		Tonnes
Gas in		
Air-G		25.28
Strong $FeCl_2$ liquor in		
HCl-AQ		12.89
H_2O-L		97.41
$FeCl_2$-AQ		61.93
	Total	172.23
	Total in	197.51
Gas and vapour out		
O_2-G		1.96
N_2-G		19.41
H_2O-G		7.91
	Total	29.28
Liquid and solids out		
HCl-AQ		12.89
H_2O-L		89.49
$FeCl_3$-AQ		52.83
Fe_2O_3-S		13.01
	Total	168.22
	Total out	197.50

PORI hydrolyzation evaporator using no. 6 fuel oil

		Tonnes
Energy in		
Air-G		173.55
No. 6 fuel oil		11.14
	Total	**184.69**
Recycle stream in		
HCl-AQ		9.25
H_2O-L		78.86
$FeCl_3$-AQ		37.95
	Total	**126.06**
Liquid and solids in		
HCl-AQ		12.89
H_2O-L		89.49
$FeCl_3$-AQ		52.83
Fe_2O_3-S		13.01
	Total	**168.22**
	Total in	**478.97**
Product stream out		
HCl-AQ		9.25
H_2O-L		64.28
$FeCl_3$-AQ		37.95
Fe_2O_3-S		39.02
	Total	**150.50**
Gas and vapour out		
HCl-G		48.52
CO_2-G		33.66
O_2-G		6.80
N_2-G		133.69
H_2O-G		105.30
	Total	**327.97**
	Total out	**478.47**

PORI hydrolyzation evaporator using natural gas

		Tonnes
Energy in		
Air-G		186.71
Natural gas		9.04
	Total	**195.75**
Recycle stream in		
HCl-AQ		9.25
H_2O-L		68.57
$FeCl_3$-AQ		37.95
	Total	**115.77**
Liquid and solids in		
HCl-AQ		12.89
H_2O-L		89.49
$FeCl_3$-AQ		52.83
Fe_2O_3-S		13.01
	Total	**168.22**
	Total in	**479.74**
Product stream out		
HCl-AQ		9.25
H_2O-L		64.28
$FeCl_3$-AQ		37.95
Fe_2O_3-S		39.02
	Total	**150.50**
Gas and vapour out		
HCl-G		48.52
CO_2-G		24.82
O_2-G		7.22
N_2-G		143.39
H_2O-G		105.30
	Total	**329.25**
	Total out	**479.75**